SOME RESULTS FROM SLAGGING, FIXED-BED GASIFICATION OF LIGNITE AT PRESSURES TO 400 PSIG

G. H. Gronhovd, A. E. Harak, and M. M. Fegley

U. S. Department of the Interior, Bureau of Mines
Grand Forks Lignite Research Laboratory
Grand Forks, N. Dak.

INTRODUCTION

Gasification of coal with steam and oxygen to produce synthesis gas is one of the promising basic starting procedures for converting our vast coal reserves into the convenient fluid fuels, gas and oil. Perhaps the most attractive gasification method, from the standpoint of thermal efficiency, is the fixed-bed system with its inherent good heat exchange between the countercurrent flowing coal and hot gases. Fixed-bed gasification is limited to those fuels that do not agglomerate under the conditions existing in the gasifier shaft. Lignite, being a nonagglomerating fuel, is well suited for this process.

In 1958 the Bureau of Mines began development of a high-pressure, fixed-bed, slagging gasification pilot plant at Grand Forks, N. Dak. The objectives of the program were to develop a suitable pilot plant technique for slagging gasification of lignite and other noncaking fuels at pressures to 400 psig and to obtain process data as a function of operating pressure and other variables. It was hoped that the new process would show significant advantages over the conventional Lurgi[1] pressure gasifier, which operates with dry ash removal. Some of the advantages predicted for slagging operation were increased capacity, increased thermal efficiency, reduced steam consumption, and reduced quantity of waste liquors. Other advantages of a slagging gasifier are its ability to use coals having low ash fusion temperatures and the absence of a mechanical grate, which can require considerable maintenance in a large commercial gasifier.

Slagging pressure gasification does, however, pose some formidable problems in refractory erosion and in methods of discharging molten slag from a pressure vessel.

The gasifier was originally equipped with a low-pressure bottom section, and the early development work was done at 80 psig. A bituminous coal char, Disco, was used as feed material for the initial tests because it was tar-free, and the char ash, when fluxed with blast-furnace slag, had very desirable flow characteristics. In general, the development procedure has been to establish a satisfactory slagging technique for a given design and pressure, using Disco char, and then, after developing necessary techniques, to switch to operation with lignite.

Based on the experience and data obtained at 80 psig, a new high-pressure bottom section was designed for the gasifier. This unit was installed in October 1962, and since then, the gasification pilot plant has been operated at pressures

[1] Trade names are used for identification only, and endorsement by the Bureau of Mines is not implied.

to 400 psig. Disco char was used for the first high-pressure tests, but in May 1963 a series of tests was started to demonstrate the operability of the gasifier at high pressures using North Dakota lignite. Since then, eight tests have been run at pressures from 200 to 400 psig, and this paper will be primarily concerned with the experiences and results from these tests.

DESCRIPTION OF THE PROCESS

A flowsheet of the slagging gasification pilot plant is presented in figure 1. The fuel, which is periodically charged to the coal lock, moves by gravity flow into the generator and is continuously gasified by an oxygen-steam mixture introduced through four water-cooled tuyères at the bottom of the gasifier. Molten slag is formed at the hearth and flows through a central 1-inch-diameter taphole into a water quench bath. The slag-water slurry is periodically discharged from the slag lock.

The product gas leaving the gasifier and containing water vapor and tar is scrubbed in the spray cooler with recycled liquor that has been condensed out of the gas. The washed gas is then cooled to about 60° F in an indirect cooler before being sampled, metered, and flared.

Some cooled product gas is compressed and recycled through the coal lock to prevent steam and tar vapors from condensing on the cold fuel.

Gas from the high-temperature reaction zone can be drawn through the taphole to aid slag flow. This gas is cooled and metered in a separate circuit.

A cross-section view of the gasifier is presented in figure 2. The inside diameter of the unit is 16-5/8 inches, giving a fuel bed area of 1-1/2 square feet. An available maximum fuel bed depth of 15 feet provides heat exchange and gas residence times similar to those in a commercial Lurgi unit.

For additional details on equipment design and operating procedure, refer to the recent Bureau of Mines Report of Investigations 6085 (1).

FUELS AND FLUXES TESTED

North Dakota lignite as-mined usually contains from 35 to 40 percent moisture. Before being charged to the gasifier, this lignite is dried to 20 to 25 percent moisture in a Fleissner steam drier and then screened to 3/4 by 1/4 inch. A typical analysis of a dried lignite as charged to the gasifier is shown in table 1, along with an analysis of Disco char.

One of the most important criteria for a coal to be used in any slagging process is that the molten slag must have a sufficiently low viscosity to flow readily at the temperatures and atmospheres developed in the process. As an aid to predicting desirable fuel and flux combinations to be used in the slagging gasifier, viscosity determinations were made by an industrial laboratory on a number of prepared samples. Figure 3 shows viscosity-versus-temperature curves for Disco char ash, for lignite ash, and for these ashes fluxed with blast-furnace slag in the ratio of 3 pounds flux per pound of ash. Chemical compositions for the various materials are given in table 2.

The Disco ash, either alone or fluxed with blast-furnace slag, exhibited essentially Newtonian flow over the temperature range investigated, and the flux was very effective in reducing the slag viscosity.

TABLE 1. - Typical analysis of fuels gasified

	North Dakota lignite	Disco char
Proximate analysis, percent:		
Moisture	22.4	2.5
Volatile matter	31.4	19.2
Fixed carbon	39.8	67.8
Ash	6.4	10.5
Ultimate analysis, percent:		
Hydrogen	5.9	3.6
Carbon	52.4	72.0
Nitrogen	0.9	1.6
Oxygen	34.0	10.4
Sulfur	0.4	1.9
Ash	6.4	10.5
Heating value, Btu/lb	8,610	12,100
Ash fusibility temperature, °F:		
Initial deformation	2,100	2,130
Softening	2,135	2,190
Fluid	2,170	2,500

TABLE 2. - Typical chemical analysis of various fuel ashes and fluxes

	Lignite ash	Lignite clinker	Disco char ash	Blast furnace slag
Analysis, percent:				
SiO_2	33.0	41.8	48.0	36.3
Al_2O_3	13.5	17.1	23.0	11.8
Fe_2O_3	7.3	8.0	18.5	3.9
TiO_2	0.6	0.8	1.0	0.3
CaO	14.5	16.9	3.3	36.6
MgO	4.4	4.4	1.0	9.0
Na_2O	12.1	8.5	0.3	0.4
K_2O	1.0	0.9	1.9	0.3
SO_3	13.8	2.6	2.7	0.9
Mn_3O_4	--	--	--	0.5
Ash fusibility, °F:				
Initial deformation	1,910	2,020	2,050	2,440
Softening	2,050	2,050	2,140	2,480
Fluid	2,140	2,140	2,430	2,520

The lignite ash had relatively low viscosity in the fluid range but exhibited a sharp deviation from Newtonian flow at about 2,000° F. This point at which the viscosity increases rapidly with a small drop in temperature is called the "temperature of critical viscosity." The addition of blast-furnace slag reduced the slag tappability of the lignite ash by increasing the "temperature of critical viscosity" by about 350° F.

The laboratory-determined ash viscosities are helpful in assessing suitability of fuel-flux combinations for slagging gasification, and we have had some success in correlating these data with actual performance in the gasifier. However, the laboratory conditions where the sample is slowly heated and allowed to come to equilibrium at each temperature are quite different from conditions in the gasifier where the slag upon formation flows rapidly to the taphole and may not react completely with the flux material. Also, the atmosphere surrounding the slag in the gasifier may be quite different from the atmosphere in the laboratory test.

In our small gasifier it is sometimes necessary to add fluxing material with certain fuels to reduce the slag viscosity, and it is usually desirable to increase the quantity of slag flowing through the taphole and thus reduce heat loss per pound of slag flowing.

GASIFICATION OF NORTH DAKOTA LIGNITE

In some of the early lignite tests, blast-furnace slag was tried as a fluxing material, but, as predicted from the laboratory viscosity data, reliable slagging operation could not be obtained. Various methods for improving the lignite slag flow were tried, but the most successful procedure has been to add sized combustion clinkers from a local powerplant to the lignite charge. By this method the chemical composition of the ash is not appreciably changed, and the slag flow can be increased to any desired rate. The ratio of clinkers to lignite for the tests to date has ranged from 30 to 10 pounds per 100 pounds of lignite, depending upon the gasification rate. The average slag flow rates have generally been from 150 to 200 pounds per hour.

The operability of the gasifier on lignite at pressures to 400 psig was generally satisfactory, although in some tests difficulties in maintaining slag flow were encountered after 5 to 10 hours of slagging operation. As will be discussed later in the report, this problem seems to be related to iron segregation on the hearth.

Contrary to expectations, no trouble has been experienced in carrying tar vapors out of the gasifier despite the low gas offtake temperatures, often less than 300° F.

TEST RESULTS

The gas production rate is determined by the oxygen-steam input rates, and during this series of eight tests, oxygen rates used were from 4,000 to 6,500 cubic feet per hour. The oxygen-steam molar ratio was maintained at approximately 0.9 for all tests. Because of limitations on gas cooling and metering equipment, the maximum capacity was limited to 6,500 cubic feet per hour oxygen rate.

Typical results from gasification of lignite at 400 psig and 6,000 cubic feet per hour oxygen input are given in table 3. For comparison, results from a Disco char test at the same pressure and oxygen rate are also given. The gas production rate is about 29,000 cubic feet per hour, or 19,300 cubic feet per square foot per hour, for both tests. The lignite feed rate is 1,442 pounds per hour; 934 pounds per hour is the rate for Disco. The gas analysis for the lignite test shows higher CH_4 and CO_2 as a result of increased pyrolysis gas from this fuel. The concentration of sulfur compounds in the product gases are approximately proportional to the total sulfur input in the fuel.

Table 4 is a typical material balance for a lignite test at 400 psig, and table 5 is a heat balance for the same test. The heat loss to the slag is 1.3 percent, and the other unaccounted for losses are 4.0 percent of the total input heat.

TABLE 3. - Typical results from gasification at 400 psig

	Fuel	
	Steam-dried lignite	Disco char
Oxygen rate............................scf/hr1/	6,000	6,000
Oxygen/steam ratio ....mole/mole	0.9	0.8
Fuel rate.............................lb/hr	1,442	934
Fuel rate (maf)......................lb/hr	987	792
Flux ratio...................lb/100 lb fuel	10	15
Total gas rate........................scf/hr	29,000	29,100
Specific gas rate.................$scf/ft^2/hr$	19,300	19,400
Slag discharge rate....................lb/hr	214	231
Material requirements per Mft^3 crude gas:		
Oxygen..............................ft^3	207	206
Steam................................lb	10.9	12.2
Fuel (maf)...........................lb	34.0	27.2
Material requirements per Mft^3 ($CO+H_2$):		
Oxygen..............................ft^3	245	227
Steam................................lb	12.9	13.5
Fuel (maf)...........................lb	40.2	30.0
Cold gas efficiency..................percent2/	84.5	89.5
Operational efficiency..................do..3/	91.0	89.5
Average gas offtake temperature............°F	508	880
Steam decomposition..................percent4/	-	83.4
Crude gas analysis:5/		
CO_2................................do..	7.5	3.3
Illuminants.............................do..	0.4	0.1
O_2...................................do..	0.1	0.2
H_2...................................do..	28.4	29.7
CO.....................................do..	56.2	61.2
C_2H_6................................do..	0.5	0.0
CH_4..................................do..	6.9	5.5
Heating value.........................Btu/ft^3	354	345
H_2S.......................$grains/100\ ft^3$	107	284
Organic sulfur..............$grains/100\ ft^3$	20	73

1/ All gas volumes at 30 in. Hg and 60° F.

2/ Cold gas efficiency $= \frac{\text{potential heat in gas}}{\text{potential heat in fuel}} \times 100.$

3/ Operational efficiency $= \frac{\text{potential heat of gas + potential heat of tar}}{\text{potential heat in fuel}} \times 100.$

4/ Steam decomposition $= \frac{H_2 \text{ in gas + tar} - H_2 \text{ in mf fuel}}{H_2 \text{ in steam}} \times 100.$

5/ Gas analysis is calculated to nitrogen-free basis to correct for inert gas added to the system, and gas production rates are given on this basis.

TABLE 4. - Typical material balance from gasification of lignite at 400 psig

	Pounds per hour	Percent
Material in:		
Steam	316	13.1
Oxygen	507	21.0
Fuel	1,442	59.9
Flux	144	6.0
Total	2,409	100.0
Material out:		
Product gas	1,701	70.6
Slag	214	8.9
Condensate + tar	495	20.5
Unaccounted for	-1	0.0
Total	2,409	100.0

TABLE 5. - Typical heat balance from gasification of lignite at 400 psig

	M Btu per hour	Percent
Heat in:		
Fuel	12,153	96.4
Steam	401	3.2
Oxygen	54	0.4
Total	12,608	100.0
Heat out:		
Product gas:		
Potential	10,273	81.5
Sensible	270	2.1
Slag	166	1.3
Water vapor	620	4.9
Tar (potential)	781	6.2
Unaccounted for losses	498	4.0
Total	12,608	100.0

Figure 4 shows the variation in gas analysis with operating pressures when gasifying lignite at pressures from 80 to 400 psig. The CO and H_2 concentrations decrease slightly with pressure, and the CO_2 and CH_4 concentrations increase slightly.

LIGNITE TAR

In a fixed-bed gasifier of this type, the moisture and pyrolysis products of the fuel are carried out of the gasifier with the product gas. These tar and water vapors are condensed in the spray washer by direct contact with cooled recycled liquor. The tar is discharged from the spray washer in the form of a tar-water emulsion that usually contains from 40 to 50 percent water. The specific gravity of the tar is very close to unity, and no satisfactory method of breaking this emulsion has been found. Commercial Lurgi plants are also plagued with this problem (2). The tar recovery has varied considerably during the test program, usually ranging from 10 to 20 gallons of moisture-free tar per ton of moisture- and ash-free lignite. The variations in yield are believed to be caused, to a large extent, by the tar separation and recovery problems.

Table 6 shows data on lignite tar obtained at operating pressures from 80 to 400 psig. In general, there appears to be little difference between the various tars.

TABLE 6. - Properties of tars and tar fractions from the pressure gasification of North Dakota lignite

Run number	51	P-15	P-16	P-20
Pressure, psig	80	200	300	400
Moisture, weight-percent:				
Distillation to cracking	51.6	52.2	55.0	42.1
Primary distillation, weight-percent of dry tar:				
Distillate	82.6	74.8	79.6	76.4
Pitch	11.8	19.6	15.5	18.3
Loss	5.6	5.6	4.9	5.3
Temperature of decomposition, °C	300	310	300	332
Composition of distillate, weight-percent:				
Tar acids	44.0	39.0	39.2	36.9
Tar bases	1.2	0.4	0.4	0.5
Neutral oil	54.8	60.6	60.4	62.6
Distillate, Hempel distillation, weight-percent:				
To 95° C	4.2	6.6	5.5	2.7
95-105	0.9	0.9	0.1	0.5
105-170	3.7	1.9	1.9	5.5
170-185	4.3	2.7	6.2	4.5
185-200	7.5	6.0	6.6	10.9
200-210	5.1	7.0	6.2	6.3
210-235	12.7	14.6	10.0	11.3
235-270	13.9	10.3	13.9	10.2
270-decomposition	30.9	40.5	39.2	39.7
Pitch	13.8	6.7	7.7	7.0
Loss	3.0	2.8	2.7	1.4
Temperature of decomposition, °C	360	370	382	376
Tar recovery, gal tar per ton maf coal	21.0	8.7	13.8	10.8
Specific gravity, 25/25 °C	1.0208	1.0570	1.0423	1.0453
Ultimate analysis, percent:				
Carbon	83.8	83.2	82.9	83.3
Hydrogen	9.1	8.4	8.6	8.4
Nitrogen	-	1.0	0.9	1.0
Oxygen	-	6.7	6.9	6.6
Sulfur	-	0.7	0.7	0.7

SIZE DEGRADATION IN THE FUEL BED

As the lignite passes down through the drying and carbonizing zones in the gasifier, significant size reduction occurs. Additional fines are probably produced in the turbulent zones of the raceways in front of each tuyère. Figure 5 shows screen analyses of the material charged to the gasifier and of the material in the upper and lower portions of the bed after a test using either lignite or Disco char. The lignite-clinker mixture as charged contains about 66 percent plus 1/2-inch material and about 2 percent minus 1/8 inch. Screen analysis of the lower 3 feet of the fuel bed at the conclusion of a test showed 4.4 percent of the plus 1/2-inch material and 54.4 percent of the minus 1/8-inch material. The deep fuel bed tends to filter out this fine material, and no fuel bed carryover has been noted at 400 psig with gasification rates as high as 20,000 cubic feet per square foot per hour. As shown, the size degradation in the Disco char tests were much less than for lignite.

Because of the greater amount of fines when gasifying lignite and the resultant greater pressure drop for a given gasification rate, it is expected that the maximum capacity will be less for lignite than for Disco char.

EFFECT OF COAL CHARGE CYCLE

During the high-capacity tests at 400 psig, the fuel flux feed rate was about 1,570 pounds per hour, or about 1,000 pounds per square foot per hour. This rate required charging the coal lock every 45 minutes. The charging operation usually required about 11 minutes, and since no fuel was being fed to the shaft during this time, the fuel bed level dropped an estimated 4 to 5 inches per minute, or a total of about 4 feet. This fluctuation in fuel bed height has a great effect on the gas offtake temperature and some effect on the gas composition. As shown in figure 6, the gas offtake temperature is 300° to 400° F during the time that coal is being fed to the shaft, but when the coal lock becomes empty, this temperature quickly rises to 1,500° to 1,600° F. The bed temperature 10 feet above the hearth is also plotted in figure 6 and shows no effect of the charge cycle but remains constant at about 2,300° F.

The CO_2 in the product gas is continuously monitored by an infrared analyzer, and these data are also plotted in figure 6. The CO_2 drops sharply during the charging period by as much as 5 percent.

The drop in fuel bed height during charging could ultimately limit the maximum capacity of the gasifier because the remaining fuel bed will contain a greater percentage of fines and will thus fluidize more easily. This problem of fuel bed height variation could be solved by adding another coal lock to the gasifier, either in series or parallel with the present unit. To do this now would require extensive modifications to the gasifier and supporting structure, and this change is not presently being considered.

SLAG FLOW PROBLEMS

As was stated earlier, some of the tests in this series proceeded smoothly for 5 to 10 hours of slagging operation, and then sudden freezing of material in the taphole caused premature shutdown. In most of these instances, several pounds of iron agglomerates were found on the hearth bottom during cleanout, and solidified iron streams indicated flow toward or inside of the taphole. It appears that the lignite slag itself is very fluid and flows well until the iron pools become large enough and start flowing toward the taphole.

Indications are that most of the metallic iron produced flows uniformly through the taphole with the slag. For example, in one test it was estimated that each coal-clinker charge contained the equivalent of about 9.5 pounds of iron. Analysis of the slag collected per charge showed 2.2 pounds of magnetic material, which would indicate that about 25 percent of the iron is being reduced. Since a coal-clinker charge produces about 2.2 pounds of iron per 45 minutes and the troublesome iron agglomerate of several pounds does not occur until after 5 to 10 hours of operation, it follows that most of the iron produced flows uniformly through the taphole with the slag. The phenomenon causing the iron agglomerates on the hearth bottom and the resultant flow problems are not understood but are currently being studied. The separated iron may have a higher melting temperature than the lignite slag, and the intimate contact between the hearth bottom and the iron may be responsible for the iron freezing before discharge.

COMPARISON OF RESULTS WITH COMMERCIAL LURGI

Detailed results from the new Westfield Lurgi gasification plant in Scotland have recently been published (3, 4), and in table 7 the preliminary high-pressure results from the experimental slagging gasifier are compared with those from the dry-ash Lurgi plant.

The Westfield plant uses a bituminous coal; however, on an as-charged basis, the volatile matter and fixed carbon contents of this fuel are about the same as for the steam-dried lignite.

Ricketts states that gas production rates of 11,000 cubic feet of crude gas per square foot per hour have been obtained at Westfield and indicates that even higher rates may be possible. This compares with a rate of 19,300 cubic feet per square foot per hour which has presently been obtained with the slagging gasifier. We believe this rate will be increased substantially in future tests. On the basis of ($CO+H_2$), the slagging gasifier rate per square foot is about 2.4 times that of the conventional Lurgi because of the much higher CO and lower CO_2 in the product gas from the slagging operation.

The material requirements per 1,000 cubic feet ($CO+H_2$) for the two processes are also given in table 7. The maf fuel requirements are 40.2 pounds for the slagging gasifier vs 47.4 pounds for the Lurgi. The oxygen requirements are 245 cubic feet vs 267 cubic feet for the Lurgi. The largest difference is in the steam requirements; the slagging gasifier uses only 12.9 pounds, compared with 67.4 pounds for the dry ash unit. The average gas offtake temperature for the Westfield gasifier is about 900° to 1,000° F, compared with 500° F for the slagging gasifier.

SUMMARY

Slagging gasification of North Dakota lignite at pressures to 400 psig has been demonstrated in a beginning series of eight tests at high pressure. Performance of the gasification pilot plant during these tests was generally good, although taphole plugging limited some tests to about 5 to 10 hours of slagging operation. This taphole freezing problem appears to be directly related to iron segregation and agglomeration on the hearth and constitutes the most serious impediment to extended operation of the gasifier at the present time. No flux was used with the lignite, and, neglecting the iron segregation, the slag was very fluid and flowed nicely through the taphole, indicating the suitability of lignite for a slagging process.

TABLE 7. - Comparison of results from slagging gasifier and commercial Lurgi plant

	Slagging gasifier	Westfield Lurgi plant
Operating pressure, psig	400	355
Fuel gasified	Steam-dried lignite	Bituminous
Fuel analysis, proximate, percent		
Moisture	22.4	16.0
Ash	6.4	14.3
Volatile matter	31.4	28.5
Fixed carbon	39.8	41.2
Gas production rate:		
Crude gas, $ft^3/ft^2/hr$	19,300	11,000
$(CO+H_2)$, $ft^3/ft^2/hr$	16,300	6,970
Gas analysis, percent:		
CO_2	7.5	26.1
O_2	0.1	-
N_2	-	0.8
CO	56.2	26.0
H_2	28.4	37.4
CH_4	6.9	9.1
C_2H_6	0.5	-
CnHm	0.4	0.6
Heating value, Btu/ft^3	354	310
Material requirements per Mft^3 $(CO+H_2)$:		
Fuel, lb:		
As-charged	58.8	68.0
Moisture- and ash-free	40.2	47.4
Fixed carbon	22.4	28.0
Steam, lb	12.9	67.4
Oxygen, ft^3	245	267
Average gas offtake temperature, °F	500	900-1,000
Cold gas efficiency,[1] percent	84.5	81.0

1/ Cold gas efficiency $= \frac{\text{Potential heat in product gas}}{\text{Potential heat in the coal}} \times 100$

Considerable process data were obtained from the lignite tests at various pressures to 400 psig. The gas composition, tar composition, and material requirements per unit synthesis gas showed only small variation with operating pressure.

The maximum gas production rate during this series was limited to about 20,000 cubic feet per square foot per hour because of limitations on gas metering and cooling equipment. However, at this rate and at 400 psig, there was no indication of fuel bed carryover. It is expected that considerably higher rates will be obtained in future tests.

Comparison of results with those of the Westfield dry-ash Lurgi plant shows favorable material requirements of the slagging gasifier per unit of synthesis gas, and production rates per square foot are more than double those from the Lurgi plant.

Because of the drop in fuel bed height while coal is being charged to the gasifier, the gas offtake temperature increases from 300° to 1,600° F and the CO_2 in the product gas decreases by about 5 percent. This fluctuation is undesirable and could be avoided either by using double lock-hoppers for the coal feed or by several other possible modifications of the fuel-charging system.

No trouble has been experienced in carrying tar vapors out of the gasifier despite the generally low gas offtake temperatures. Difficulties in tar-water emulsion separation are blamed for inconsistent results in tar yields per ton of lignite, which range from 10 to 20 gallons of moisture-free tar per ton of moisture- and ash-free lignite.

Size degradation in the shaft is quite severe when using lignite, and over 50 percent of the material in the lower part of the shaft is minus 1/8 inch. The deep fuel bed acts as an efficient filter to prevent dust carryover; however, the additional pressure drop caused by the fine material should cause hangup of the fuel bed at lower gas production rates with lignite than with Disco char.

The hearth design as developed for this small pilot gasifier would not be suitable for scale-up for a commercial unit. The life expectancy of the refractory taphole is less than 50 hours, and some type of water-cooled hearth and taphole is envisioned for a large gasifier. In addition, some means of heating the taphole during periods of low load or temporary shutdown would probably be required. The British investigators have made significant progress in the development of these items (5, 6).

The pilot plant is now being modified to permit higher capacity operation, and tests will be continued in order to determine maximum capacity and other process data as a function of operating pressure and other variables.

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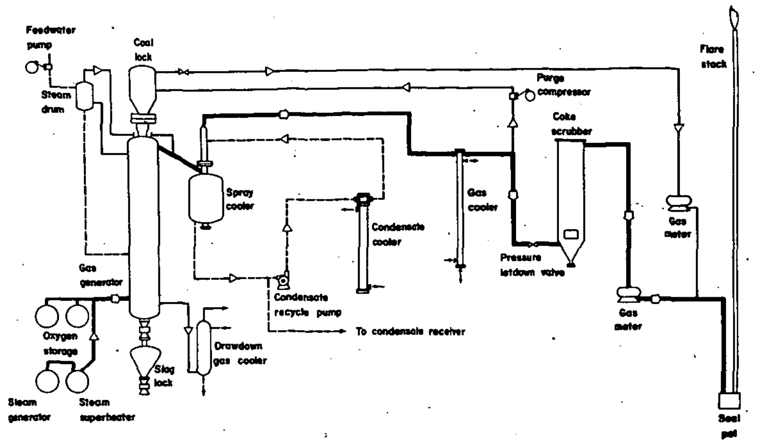

Fig. 1 Process flow diagram for pressure gasification pilot plant.

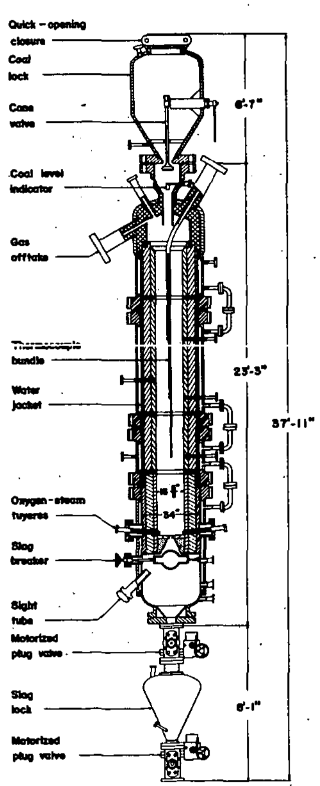

Fig. 2 Cross section of pressure gasifier.

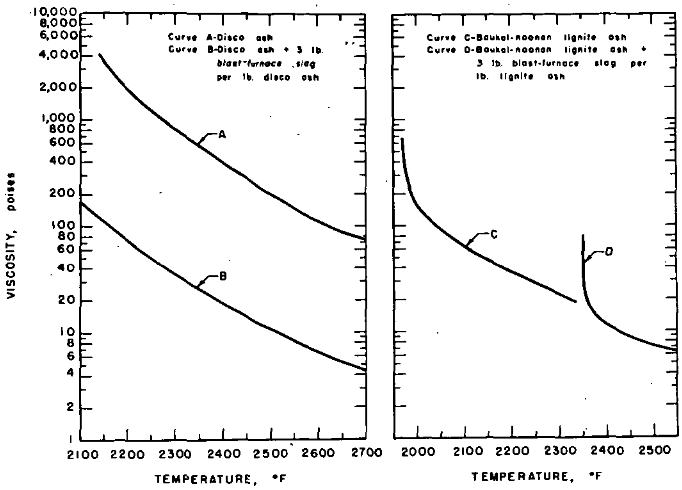

Fig. 3 Measured viscosity-temperature relationships for various coal ashes and for mixture of coal ashes and flux.

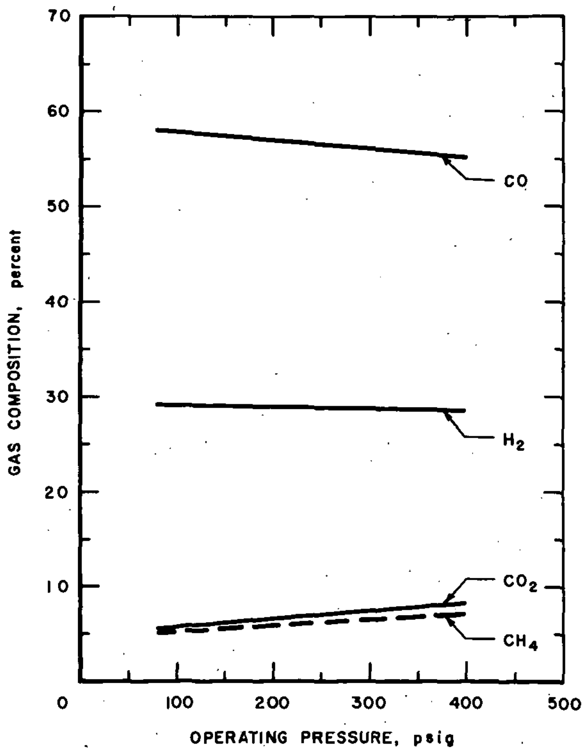

Fig. 4 Gas analysis versus operating pressure using North Dakota lignite.

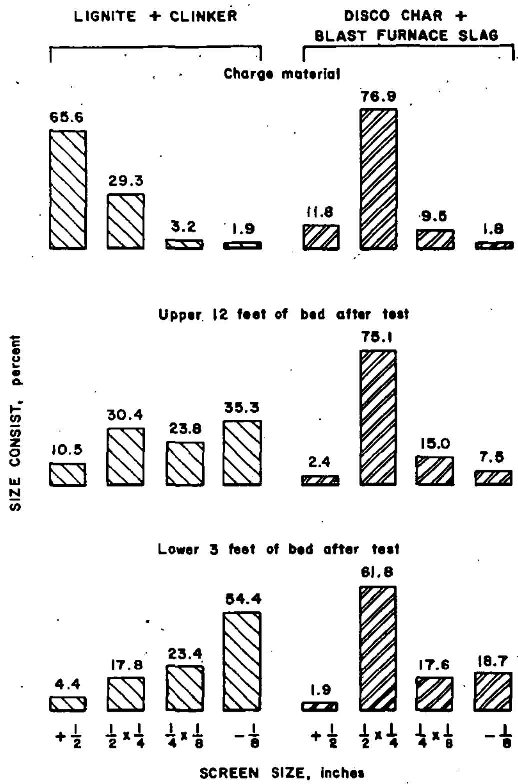

Fig. 5 Typical screen analysis of material fed to gasifier and fuel bed removed from gasifier after test.

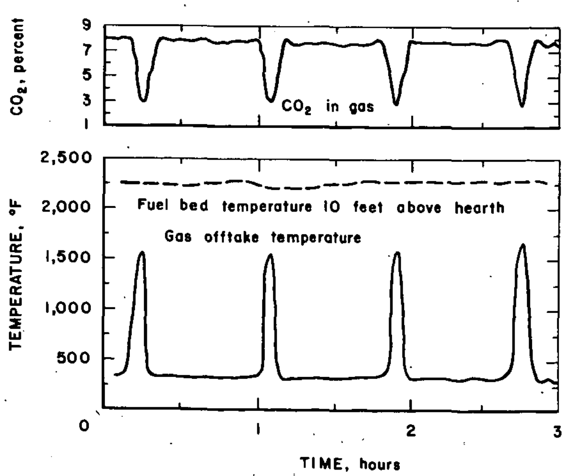

Fig. 6 Gas offtake temperature, fuel bed temperature, and CO_2 in product gas as influenced by the coal charge cycle.